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APPLICATION NUMBER: *60/517,298*
FILING DATE: *November 03, 2003*
RELATED PCT APPLICATION NUMBER: *PCT/US04/19490*

By Authority of the
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17414 U.S.PTO 110303

Practitioner's Docket No. 100325.0236PRO

PATENT

8

Preliminary Classification Proposed Class: Subclass:

IN THE UNITED STATES PATENT AND TRADEMARK OFFICE

22387 U.S.PTO 60/517298

In re application of: John Mak, Richard B. Nielsen and Curt Graham

For: LNG Vapor Handling And Regasification Methods And Configurations

Mail Stop Provisional Patent Application
Commissioner for Patents
P.O. Box 1450
Alexandria, VA 22313-1450

COVER SHEET FOR FILING PROVISIONAL APPLICATION
(37 C.F.R. § 1.51(c)(1))

This is a request for filing a PROVISIONAL APPLICATION FOR PATENT under 37 C.F.R. § 1.51(c)(1)(i). The following comprises the information required by 37 C.F.R. § 1.51(c)(1):

1. The following comprises the information required by 37 C.F.R.§ 1.51(c)(1):

2. The names of the inventors are (37 C.F.R. § 1.51(c)(1)(ii)):

   1. John Mak
      2242 Salt Air Drive
      Santa Ana, CA 92705

EXPRESS MAILING UNDER 37 C.F.R. § 1.10*

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I hereby certify that this paper, along with any document referred to, is being deposited with the United States Postal Service on this date November 3, 2003 in an envelope addressed to the Commissioner for Patents, P.O. Box 1450, Alexandria, VA 22313-1450 as "Express Mail Post Office to Addressee" Mailing Label No. EL961800551US.

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2. Richard B. Nielsen
18 Callender Court
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3. Curt Graham
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Mission Viejo, CA 92691

3. The title of the invention is (37 C.F.R. § 1.51(c)(1)(iv)):

LNG Vapor Handling And Regasification Methods And Configurations

4. The name, registration, customer and telephone numbers of the practitioner are (37 C.F.R. § 1.51(c)(1)(v)):

Name of practitioner: Robert D. Fish
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Tel. 714-641-5100
Customer No. 34284

5. The docket number used to identify this application is (37 C.F.R. § 1.51(c)(1)(vi)):

Docket No. 100325.0236PRO

6. The correspondence address for this application is (37 C.F.R. § 1.51(c)(1)(vii)):

Rutan & Tucker, LLP
611 Anton Blvd., Suite 1400
Costa Mesa, CA 92626

7. Statement as to whether invention was made by an agency of the U.S. Government or under contract with an agency of the U.S. Government. (37 C.F.R. § 1.51(c)(1)(viii)).

This invention was NOT made by an agency of the United States Government, or under contract with an agency of the United States Government.

8. Identification of documents accompanying this cover sheet:

A. Documents required by 37 C.F.R. § 1.51(c)(2)-(3):

Specification:	No. of pages	8
Drawings:	No. of sheets	4

B. Additional documents:

9. Fee

The filing fee for this provisional application, as set in 37 C.F.R. § 1.16(k), is $160.00 for other than a small entity.

10. Fee payment

Fee payment in the amount of $160.00 is being made at this time.

11. Method of fee payment

Charge Account No. 502191, in the amount of $160.00.

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Date: 11/3/03

Robert D. Fish
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LNG VAPOR HANDLING AND REGASIFICATION METHODS AND CONFIGURATIONS

This application claims the benefit of U.S. provisional patent application number with the serial number 60/476,770, filed June 5, 2003, and which is incorporated herein by reference.

Field of the Invention

The field of the invention is gas processing, especially as it relates to vapor handling during LNG ship unloading and LNG regasification and processing of liquefied natural gas.

Background of The Invention

With increasing natural gas demand in the United States, import of liquefied natural gas (LNG) has gained considerable momentum. Major expansion and additions of new LNG receiving and regasification terminals are in various phases of design, engineering and construction. In respect to the design of an LNG import terminal, LNG ship unloading is a critical operation that must be efficiently integrated with the regasification operation, particularly in the handling the vapor evolution during the unloading operation. When LNG is unloaded from the LNG ship to the storage tank, vapor is generated from the storage tank that must be properly handled and recovered to avoid flaring and pressure buildup in the storage tank system. The vapor generation is attributed to several factors including volumetric displacement, the heat gain to LNG transfer and pumping system, storage tank boiloff, and flashing due to the pressure differential between the ship and the storage tank. In a typical LNG receiving terminal, a portion of the vapor is returned to the LNG ship, and the remaining portion is compressed by a compressor that increases the pressure sufficiently for condensation in a vapor absorber using the refrigeration content from the LNG sendout. The vapor compression and vapor absorption system requires significant energy consumption and operator attention particularly the during transition from the normal holding operation (when the ship unloading facility is idled awaiting for next ship) to the ship unloading operation. The vapor boiloff from the storage tank is significantly less during the holding operation,

and, in some facilities, a separate smaller compressor is installed for energy conservation or, alternatively, the larger unloading compressor can be operated at turndown condition at the expense of higher energy consumption. In either option, the use of a vapor compression system for vapor handling and recovery is costly to operate and maintain, and energy inefficient.

In addition, most import LNG has a higher heating value and is richer in heavier hydrocarbons than is allowed by today's environmental standards. The higher hydrocarbons, such as C_5 and heavy components, are undesirable and damaging to the environment, in creating the undesirable CO_2 pollutants. In the past, some countries generally may accept the use of richer LNG due to its lower processing costs, the current environmental concerns, especially in North America and European markets, may dictate the new facilities to comply with the more stringent environmental regulations that may also be extended to other industrial usages. Most LNG sources (*e.g.*, from the Middle East regions or South East Asia) are generally higher in heating values and contain more $C_2 – C_6$ components, and these heavier components must be removed so as to meet the specifications on heating values, the Wobbe Index and the composition specifications.

Table 1 shows typical California pipeline gas standards and typical ranges of import LNG compositions

Component	California Standards	Typical LNG Import
C_1	88% minimum	86 to 95%
C_2	6% maximum	4 to 14%
C_3-C_5	3% maximum	3 to 7%
C_6+	0.2% maximum	0.5 to 1%
N_2 +CO_2	1.4 to 3.5%	0.1 to 1%
Gross Heating Value, Btu/SCF	970 – 1150	1050 -1200

As environmental regulations become more stringent, tighter control on LNG compositions than the current specifications are expected in the North American markets requiring new processes that can economically remove the C_2+ components from LNG. Moreover, such processes should advantageously provide a plant with sufficient flexibility to handle a wide range of LNG allowing importers to buy LNG from various low cost markets instead of being limited to those sources that meet the North America specifications.

Therefore, while the current practices and the numerous processes and configurations for LNG ship unloading and regasification are known in the art, almost all of them suffer from one or more disadvantages. Most notably, many of the currently known practices require vapor compression and absorption that are energy inefficient and are incapable to economically remove the heavy hydrocarbons from LNG in meeting stringent environmental standards. Thus, there is still a need to provide improved configurations and methods for gas processing in LNG unloading and regasification terminals.

Summary of the Invention

The present invention is directed to configurations and methods of vapor handling during LNG ship unloading and LNG regasification in a plant in which the heavy components (i.e. C_3 and heavier components) extracted from LNG are used for the absorption of vapors generated during ship unloading. Especially preferred configurations further include a fractionation system that utilizes the refrigeration released in the regasification process for the separation of LNG into a leaner natural gas and a LPG (Liquefied Petroleum Gas) product. The invention matters apply to on-shore and off-shore LNG regasification terminals.

Brief Description of The Drawing

Figure 1 is a schematic diagram of a conventional LNG ship unloading and regasification system.

Figure 2 is a schematic of one exemplary plant configuration according to the inventive subject matter.

Figure 3 is a schematic of another exemplary plant configuration according to the inventive subject matter.

Figure 4 is a schematic of one exemplary plant configuration according to the inventive subject matter for the production of a lean natural gas and LPG.

Detailed Description

The inventor has discovered that LNG can be processed in a manner that effectively utilizes its inherent refrigeration for vapor absorption during LNG ship unloading. More specifically, the inventor discovered that heavier components in LNG stream can be separated and used for absorption of vapor, eliminating the needs for compressor and vapor absorber in the conventional system. Optionally, the absorption system can be used to absorb all the vapors generated during ship unloading, eliminating the need for the conventional vapor return system. The inventive matter also includes a fractionation configuration that utilizes the refrigeration content of LNG for the production of a leaner natural gas and a LPG product.

An exemplary of a conventional LNG carrier unloading and regasification terminal is schematically depicted in **Figure 1**. LNG typically at -255°F to -260°F is unloaded from the LNG carrier via unloading arm 51, the transfer line 1 into storage tank 52, typically at a flow rate of 40,000 GPM to 60,000 GPM. The unloading operation generally lasts for about 12 to 16 hours, and during this period, about 40 MMscfd of vapor is generated from the storage tank, as a result from the enthalpy gain (either by the ship pumps or heat gain from the surroundings) during the transfer operation, the displacement vapor from the storage tanks, and the liquid flashing from the pressure difference between the ship and the storage tank. LNG ship typically operates at a pressure slightly less than that of the storage tank, and typically, the LNG ship operates at 16.2 psia to 16.7 psia while the storage tank operates at 16.5 psia to 17.2 psia. The vapor

from the storage tank, stream 2, is split into two portions, stream 3 and stream 4. Stream 3 typically at a flow rate of 20 MMscfd is returned to the LNG ship via a vapor return line and return arm 54 for replenishing the displaced volume from ship unloading. Stream 4 typically at a flow rate of 20 MMscfd is compressed by compressor 55 to about 80 psia to 115 psia and fed to the vapor absorber where the vapor is de-superheated, condensed and absorbed by the sendout LNG. The power consumption by compressor is typically 1,000 HP to 2,000 HP, depending on the vapor flow rate and compressor discharge pressure.

LNG from the storage tank is pumped by the in-tank primary pumps 53 to about 115 to 150 psia forming stream 6, at a typical sendout rate of 250 MMscfd to 1,200 MMscfd. Stream 6, a subcooled liquid at -255°F to -260°F, is routed to the absorber 58 to mix with the compressor discharge stream 5 using a heat transfer contacting device such as trays and packing. The operating pressures of the vapor absorber and the compressor are determined by the LNG sendout flow rate. A higher LNG sendout rate with a higher refrigeration content would lower the absorber pressure, and hence a smaller compressor. However, the absorber design should also consider the normal holding operation when the vapor rate is lower, and the liquid rate must be reduced to a minimal. Stream 6 is split into stream 7 and stream 8 using the respective control valve 56 and 57, as needed for controlling the vapor condensation process. The vapor absorber produces a bottom stream 9 typically at about -200°F to -220°F, which is then mixed with stream 8 forming streaming 10. Stream 10 is pumped by the secondary pump 59 to typically 1000 psig to 1500 psig forming stream 11 which is then heated in LNG vaporizers to about 40°F to 60°F as needed to meet the pipeline specifications. The LNG vaporizers are typically open rack type exchangers using seawater, fuel-fired vaporizers, or vaporizers using a heat transfer fluid.

The invention shown in **Figure 2** is a more effective vapor handling method for LNG ship unloading that is operationally coupled to an LNG regasification/processing plant. Here, vapor absorption is carried out at the storage tank overhead pressure using a heavy hydrocarbon liquid (i.e. C_3 and heavier components) for absorption, with the heavy hydrocarbon separated from LNG using a fractionator. The refrigeration content in

the LNG is used for cooling in the absorption process as well as in supplying the reflux condensing duty in the fractionator.

More particularly, the LNG ship unloading operation is similar to a conventional installation; with the exception that the compressor and vapor absorber are no longer required and are replaced by a low pressure condenser exchanger and pumping system.

The preferred inventive matter includes altering the composition of the vapors from the storage tank by mixing these vapors with a subcooled heavy hydrocarbon stream. The addition of heavy hydrocarbons increases the boiling point temperature, making condensation with LNG possible. The mixture is then pumped and separated in a subsequent fractionator for recovery and recycle of the heavy hydrocarbons.

During ship unloading, vapor from the storage tank, stream 2, is split into stream 3 and stream 4. Stream 3 typically at a flow rate of 20 MMscfd is returned to the LNG ship via a vapor return line and return arm 54 for replenishing the displaced volume from ship unloading. Stream 4 typically at a flow rate of 20 MMscfd is mixed with the heavy hydrocarbon stream 16 (containing C_4 and heavier hydrocarbons). To raise the boiling point of the mixture, typically about 200 GPM to 500 GPM heavy hydrocarbons is required from the downstream fractionation system. The mixture is cooled and condensed in exchanger 61 using the refrigeration content from the LNG stream 6 forming stream 18 typically at -240°F to -255°F. It should be appreciated that the heavy hydrocarbon composition and flow rate can be adjusted in the fractionator as necessary to absorb the vapors from the storage tank during the ship unloading and the normal holding operation.

In exchanger 61, stream 6 is heated from -255°F to about -240°F supply the necessary cooling for condensing the mixture 17. The condensate stream 18 is then pumped by pump 62 to about 120 psia to 170 psia forming stream 19. Prior to feeding the fractionator 64, the mixture is heated and partially vaporized in exchanger 63 to about -10°F to 150°F by cross exchange with the bottom liquid from the fractionator. The fractionation, operating at about 100 psia to 150 psia, separates the heavy from the mixture producing an overhead liquid stream 22 (containing C_2 and lighter components) and a bottom liquid stream 21(containing C_3 and heavier components). The fractionator is

refluxed with the refrigeration from the LNG stream 17 in an overhead condenser 65 that is integral to the fractionator. Alternatively, the reflux exchanger can be located externally to the fractionator. The fractionator is reboiled using an external heat source stream 24 with a fired reboiler, steam or other heat source.

The overhead stream 22 which is depleted of the heavy hydrocarbons is mixed with the LNG stream 23 forming stream 10. The combined stream is pumped by the secondary pump 59 to typically 1000 psig to 1500 psig forming stream 11 which is then heated in LNG vaporizers to about 40°F to 60°F as needed to meet the pipeline specifications. The LNG vaporizers are typically open rack type exchangers using seawater, fuel-fired vaporizers, or vaporizers using a heat transfer fluid.

The invention shown in **Figure 3** is based on the same vapor handling method described in the previous invention that is operationally coupled to an LNG regasification/processing plant. Here, the total vapor stream 2 is mixed with the heavy hydrocarbon stream 16, absorbed and condensed with LNG stream 6. The flow rate of stream 16 is increased correspondingly to about 400 GPM to 1,200 GPM, as needed for the absorption of the higher vapor flow. In this configuration, no vapor is returned to the ship from the storage tank to the ship during unloading, thereby the vapor return line and vapor return arm 54 can be eliminated. Instead the vapor required by the ship for maintaining volumetric balance can be generated with a small vaporizer located close to the ship. A small stream 30 at a flow rate of 20 MMscfd is vaporized in heat exchanger 67 to produce vapor stream 3 that is used for replenishing the displaced volume from the ship. The heat source to the vaporizer can be seawater or ambient air. This configuration will result in significant cost savings in the terminal design particularly in facility where there is a great distance between the ship and the storage tank.

The invention shown in **Figure 4** is enhanced for additional production of LPG and for heating value control of the LNG that may be required to meet environmental regulations. In this configuration, the overhead condenser 65 includes a second refrigeration coil 66 that uses the high pressure LNG to provide additional cooling as needed for higher reflux duty required for LPG production.

The LNG stream 26 exiting the condenser coil 65 at about -220°F to -240°F is split into two portions; stream 23 and stream 24. The split proportion varies between 0 to 100%, depending on the desirable LPG production; increasing stream 24 increases LPG production. With increasing LPG production, the distillate becomes leaner in composition resulting in the production of a LNG with less heating value that may be desirable for meeting environmental regulation.

Stream 24 is fed to the mid section of the fractionator which produces a bottom LPG stream 28, and an overhead distillate liquid stream 22 that is depleted of the heavy hydrocarbons. The distillate stream 22 is then mixed with the LNG stream 23 forming stream 10 typically at -220°F to -230°F that is further pumped by the secondary pump 59 to about 1,000 psig to 1,400 psig forming stream 11. The high pressure LNG stream is heated exchanged with the overhead vapor and is heated the reflux condenser coil 66 forming stream 27, typically at about -180°F to -200°F. Stream 27 is further heated in vaporizer 60 to meet the pipeline gas requirement. The bottom stream 28 is split into two portions; stream 25 and stream 21. Stream 21 is recycled back to exchanger 20 prior to be used for vapor absorption, and stream 25 can be sold as the LPG product.

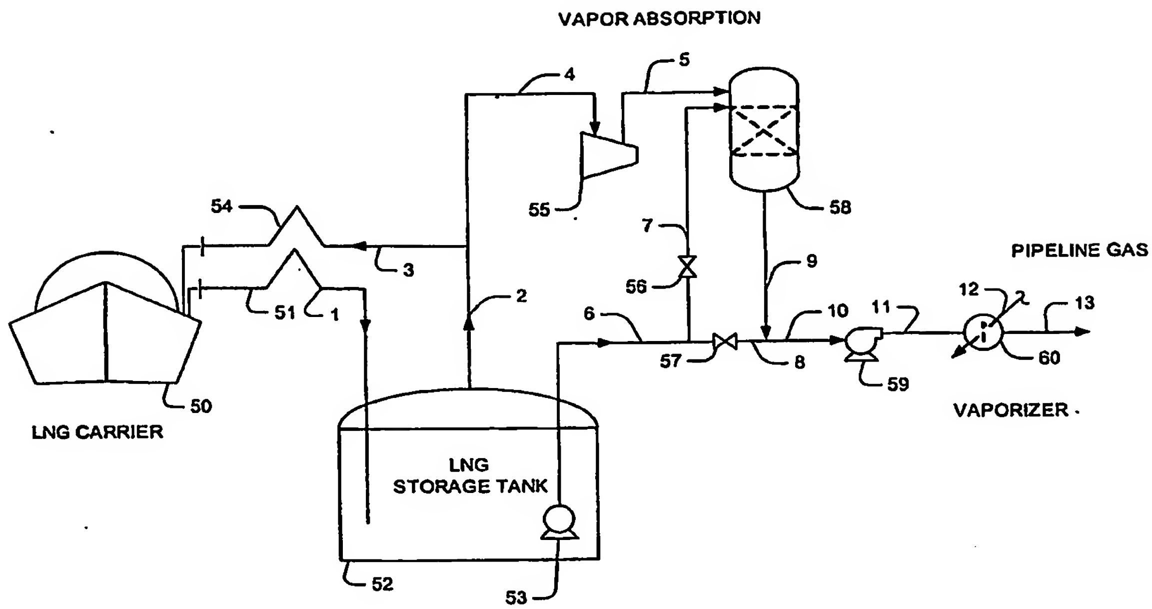
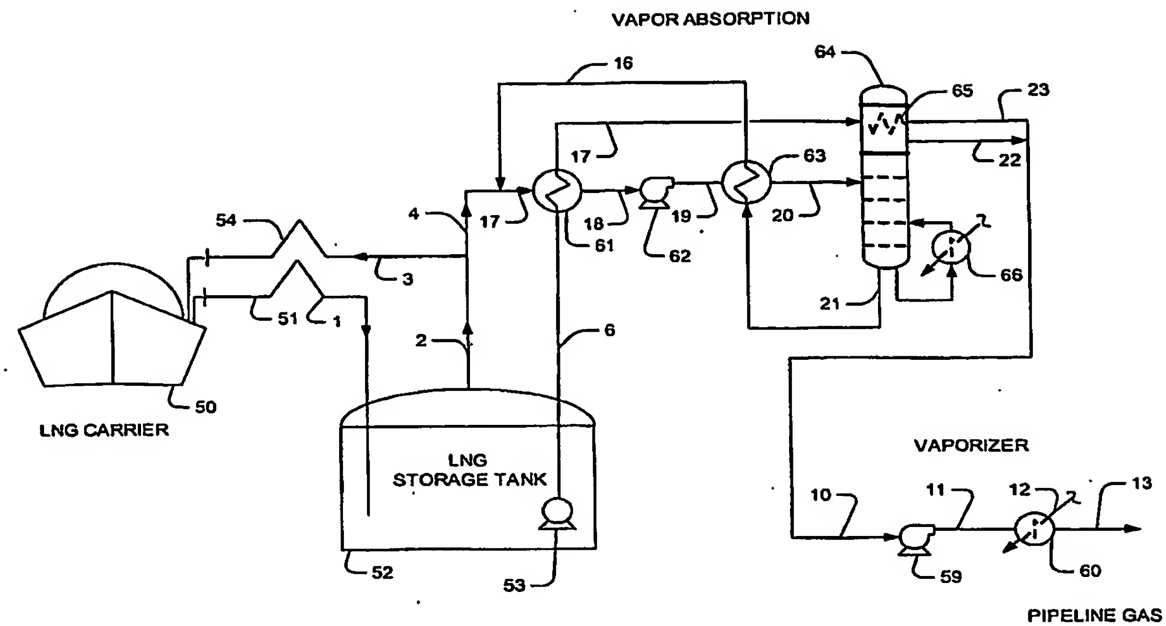


Figure 1 (Prior Art)


VAPOR ABSORPTION
64
65
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16
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22
54
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18
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3
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51
1
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6
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LNG CARRIER
LNG
STORAGE TANK
VAPORIZER
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59
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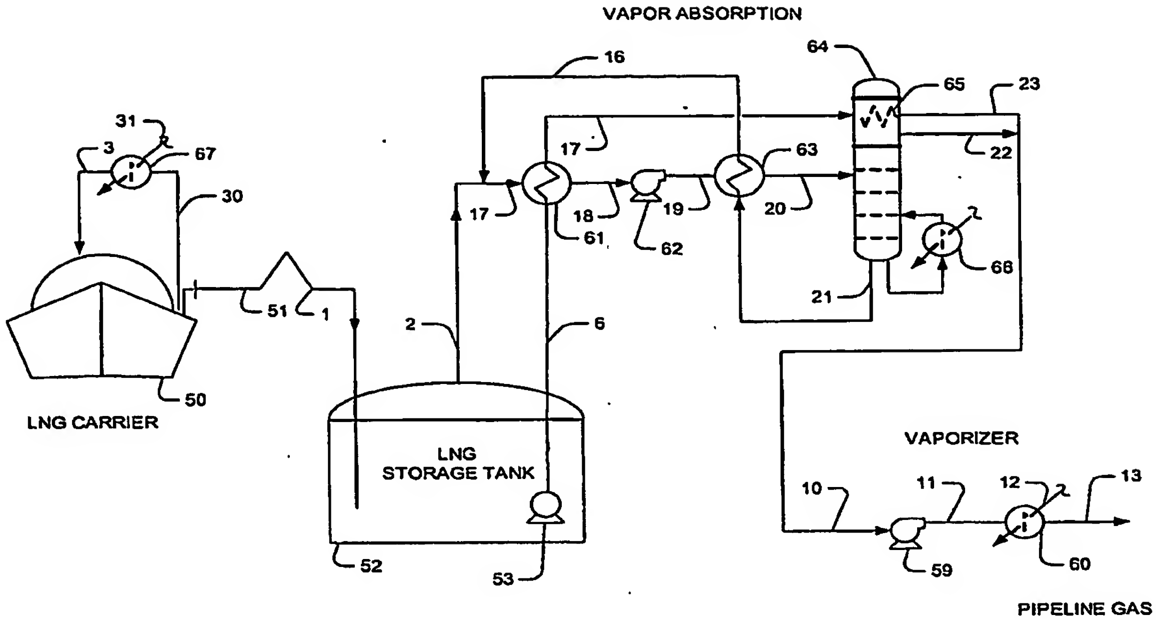
PIPELINE GAS

Figure 2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6
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LNG CARRIER
LNG
STORAGE TANK
VAPORIZER
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PIPELINE GAS

Figure 3

Atty: Docket: 100325.0236PRO

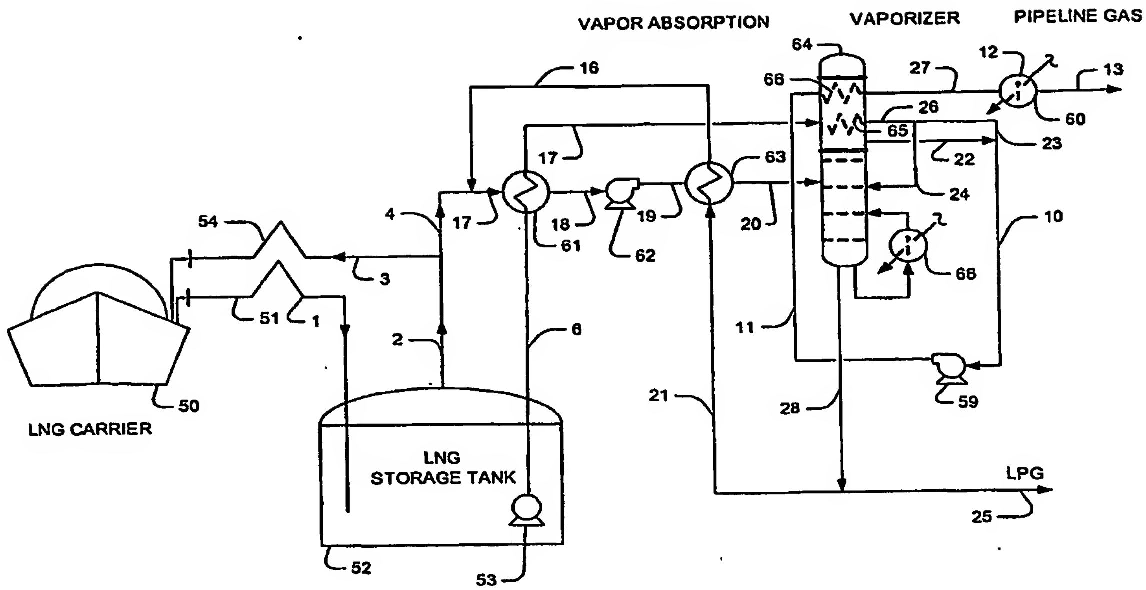

Figure 4